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APPLICATION NUMBER: 60/525,416

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RELATED PCT APPLICATION NUMBER: PCT/US04/19490

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COMMISSIONER OF PATENTS AND TRADEMARKS



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U.S. DEPARTMENT OF COMMERCE
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FEE RECORD SHEET

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Practitioner's Docket No. 100325.0240PRO

PATENT

Preliminary Classification
Proposed Class:
Subclass:

17510 U.S. PTO
60/525416

IN THE UNITED STATES PATENT AND TRADEMARK OFFICE

In re application of: John Mak, Richard B. Nielsen and Curt Graham

For: LNG Vapor Handling And Regasification

Mail Stop Provisional Patent Application
Commissioner for Patents
P.O. Box 1450
Alexandria, VA 22313-1450

COVER SHEET FOR FILING PROVISIONAL APPLICATION
(37 C.F.R. § 1.51(c)(1))

This is a request for filing a PROVISIONAL APPLICATION FOR PATENT under 37 C.F.R. § 1.51(c)(1)(i). The following comprises the information required by 37 C.F.R. § 1.51(c)(1):

1. The following comprises the information required by 37 C.F.R. § 1.51(c)(1):
2. The names and residence information of the inventors are (37 C.F.R. § 1.51(c)(1)(ii)):
 1. John Mak - Santa Ana, CA
 2. Richard B. Nielsen - Laguna Niguel, CA
 3. Curt Graham - Mission Viejo, CA

EXPRESS MAILING UNDER 37 C.F.R. § 1.10*

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I hereby certify that this paper, along with any document referred to, is being deposited with the United States Postal Service on this date November 25, 2003 in an envelope addressed to the Commissioner for Patents, P.O. Box 1450, Alexandria, VA 22313-1450 as "Express Mail Post Office to Addressee". Mailing Label No. EV389269085US.

Date: 11/25/03

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Cover Sheet for Filing Provisional Application—page 1 of 3

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3. The title of the invention is (37 C.F.R. § 1.51(c)(1)(iv)):

LNG Vapor Handling And Regasification

4. The name, registration, customer and telephone numbers of the practitioner are (37 C.F.R. § 1.51(c)(1)(v)):

Name of practitioner: Robert D. Fish
Reg. No. 33880
Tel. 714-641-5100
Customer No. 34284

5. The docket number used to identify this application is (37 C.F.R. § 1.51(c)(1)(vi)):

Docket No. 100325.0240PRO

6. The correspondence address for this application is (37 C.F.R. § 1.51(c)(1)(vii)):

Rutan & Tucker, LLP
611 Anton Blvd., Suite 1400
Costa Mesa, CA 92626

7. Statement as to whether invention was made by an agency of the U.S. Government or under contract with an agency of the U.S. Government. (37 C.F.R. § 1.51(c)(1)(viii)).

This invention was NOT made by an agency of the United States Government, or under contract with an agency of the United States Government.

8. Identification of documents accompanying this cover sheet:

A. Documents required by 37 C.F.R. § 1.51(c)(2)-(3):

Specification:	No. of pages	8
Drawings:	No. of sheets	4

9. Fee

The filing fee for this provisional application, as set in 37 C.F.R. § 1.16(k), is \$160.00 for other than a small entity.

10. Fee payment

Fee payment in the amount of \$160.00 is being made at this time.

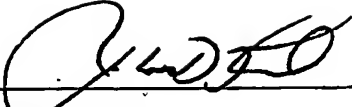
11. Method of fee payment

Charge Account No. 502191, in the amount of \$160.00.

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Date: 11/25/03


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LNG VAPOR HANDLING AND REGASIFICATION METHODS AND CONFIGURATIONS

Field of the Invention

The field of the invention is gas processing, especially as it relates to vapor handling during LNG ship unloading, LNG regasification, and processing of liquefied natural gas.

Background of The Invention

With increasing natural gas demand in the United States, import of liquefied natural gas (LNG) has gained considerable momentum. Major expansion and additions of new LNG receiving and regasification terminals are in various phases of engineering and construction.

LNG ship unloading (and especially vapor evolution) is often a critical operation that requires efficient integration with the regasification operation. In most cases, when LNG is unloaded from the LNG ship to the storage tank, vapor is generated from the storage tank and must be properly handled and recovered to avoid flaring and pressure buildup in the storage tank system. Vapor generation is generally attributed to several factors, including volumetric displacement, heat gain during LNG transfer and in the pumping system, storage tank boiloff, and flashing due to the pressure differential between the ship and the storage tank.

In a typical LNG receiving terminal, a portion of the vapor is returned to the LNG ship, while the remaining vapor portion is compressed by a compressor for condensation in a vapor absorber that uses the refrigeration content from the LNG sendout. Therefore, vapor compression and vapor absorption systems require significant energy and operator attention, and particularly during transition from normal holding operation (when the ship unloading facility is idled awaiting for next ship) to ship unloading operation. The vapor boiloff from the storage tank is significantly less during the holding operation, and in some cases, a separate costly smaller compressor is installed for this operation. Alternatively the larger unloading compressor can be operated at turndown condition at

the expense of higher energy consumption and equipment life. In either case, the use of a vapor compression system for vapor handling and recovery is complex and costly to operate and maintain.

Unfortunately, the composition and heating values of most imported LNG varies dramatically and will generally depend on the particular source. While LNG with heavier contents or higher heating value can be produced at lower costs at the source, they are not suitable for the North American market or the environmentally-conscious markets. For example, natural gas for the Californian market must meet a heating value specification of 950 Btu/SCF-1150 Btu/SCF, and must meet composition limitations on its C₂ and C₃+ components. Especially where LNG is used directly as a transportation fuel, the C₂+ content must be further reduced to avoid high combustion temperature and reduce emissions. Table 1 below compares a typical natural gas specification in North America to typical imported LNG supply sources:

COMPONENT	NATURAL GAS SPEC.	TYPICAL LNG SUPPLY
C ₁	88% minimum	87% to 94%
C ₂	6% maximum	3% to 7%
C ₃ -C ₅	3% maximum	1% to 7%
C ₆ +	0.2% maximum	0.1% to 0.8%
N ₂	1.4 to 4.5%	Less than 0.1%
Gross Heating Value, Btu/SCF	970 -- 1150	1050 - 1200

Table 1

As environmental regulations become even more stringent, tighter controls on the composition of LNG for distribution in the North American markets are expected. Therefore, new processes that can economically remove C₂+ components from LNG for these markets are needed. Moreover, such processes should advantageously provide a plant with sufficient flexibility to unload LNG from a wide range of sources and carriers, while meeting the natural gas market specifications.

Therefore, while the current practices and the numerous processes and configurations for LNG ship unloading and regasification are known in the art, almost all of them suffer from one or more disadvantages. Most notably, many of the currently known practices require vapor compression and absorption that are energy inefficient and are incapable of economically removing the heavy hydrocarbons from LNG to meet stringent environmental standards. Thus, there is still a need to provide improved configurations and methods for gas processing in LNG unloading and regasification terminals.

Detailed Description

The present invention is directed to configurations and methods of vapor handling during LNG ship unloading and LNG regasification in a plant (*e.g.*, in on-shore and/or off-shore LNG regasification terminals) in which the compression requirement for condensation of the vapors is eliminated by mixing the vapor with a component that increases the boiling point of the mixture to a degree such that at least a portion of the mixture can be condensed using the refrigeration content of the liquid portion of the LNG. It is generally further contemplated that the configurations presented herein may be implemented in both on-shore and off-shore LNG regasification terminals.

Most preferably, the admixed component comprises a heavy component (*e.g.*, C₃ and heavier components) that may be added from an external source, or more preferably, that is extracted from the LNG that is unloaded. Thus, especially preferred configurations include a fractionation system that utilizes the refrigeration released in the regasification process for the separation of LNG into a leaner natural gas and a LPG (Liquefied Petroleum Gas) product. Further especially contemplated configurations and methods for regasification of LNG that may be used in conjunction with the teachings presented herein are described in our copending U.S. provisional patent application number with the serial number 60/476,770, filed June 5, 2003, and which is incorporated herein by reference.

Configurations and methods of the inventive subject matter are contrasted with a conventional LNG carrier unloading and regasification terminal that is schematically

depicted in prior art Figure 1. Here, LNG typically at -255°F to -260°F is unloaded from the LNG carrier via unloading arm 51, the transfer line 1 into storage tank 52, typically at a flow rate of 40,000 GPM to 60,000 GPM. The unloading operation generally lasts for about 12 to 16 hours, and during this period, about 40 MMscfd of vapor is generated from the storage tank, as a result from the enthalpy gain (either by the ship pumps or heat gain from the surroundings) during the transfer operation, the displacement vapor from the storage tanks, and the liquid flashing from the pressure difference between the ship and the storage tank. An LNG ship typically operates at a pressure slightly less than that of the storage tank, and typically, the LNG ship operates at 16.2 psia to 16.7 psia while the storage tank operates at 16.5 psia to 17.2 psia. The vapor from the storage tank, stream 2, is split into two portions, stream 3 and stream 4. Stream 3 typically at a flow rate of 20 MMscfd is returned to the LNG ship via a vapor return line and return arm 54 for replenishing the displaced volume from ship unloading. Stream 4 typically at a flow rate of 20 MMscfd is compressed by compressor 55 to about 80 psia to 115 psia and fed to the vapor absorber where the vapor is de-superheated, condensed and absorbed by the sendout LNG. The power consumption by compressor 55 is typically 1,000 HP to 2,000 HP, depending on the vapor flow rate and compressor discharge pressure.

LNG from the storage tank is pumped by the in-tank primary pumps 53 to about 115 to 150 psia forming stream 6, at a typical sendout rate of 250 MMscfd to 1,200 MMscfd. Stream 6, a subcooled liquid at -255°F to -260°F , is routed to the absorber 58 to mix with the compressor discharge stream 5 using a heat transfer contacting device such as trays and packing. The operating pressures of the vapor absorber and the compressor are determined by the LNG sendout flow rate. A higher LNG sendout rate with a higher refrigeration content would lower the absorber pressure, and hence a smaller compressor. However, the absorber design should also consider the normal holding operation when the vapor rate is lower, and the liquid rate must be reduced to a minimal. Stream 6 is split into stream 7 and stream 8 using the respective control valve 56 and 57, as needed for controlling the vapor condensation process. The vapor absorber produces a bottom stream 9 typically at about -200°F to -220°F , which is then mixed with stream 8 forming stream 10. Stream 10 is pumped by the secondary pump 59 to typically 1000 psig to 1500 psig forming stream 11 which is then heated in LNG vaporizers to about 40°F to

60°F as needed to meet the pipeline specifications. The LNG vaporizers are typically open rack type exchangers using seawater, fuel-fired vaporizers, or vaporizers using a heat transfer fluid.

In contrast, as depicted in Figure 2, more effective vapor handling configurations and methods are presented in which LNG ship unloading is operationally coupled to an LNG regasification/processing plant. Here, vapor absorption is carried out at the storage tank overhead pressure using a heavy hydrocarbon liquid (e.g., C₃ and heavier) for absorption, with the heavy hydrocarbon separated from LNG using a fractionator. The refrigeration content in the LNG is used for cooling in the absorption process as well as in supplying the reflux condensing duty in the fractionator. As the mixture of the vapors and the heavy hydrocarbon liquid condenses at significantly higher temperature, it should be appreciated that a compressor and vapor absorber as depicted in prior art Figure 1 are no longer required and are replaced by a low pressure condenser exchanger and pumping system.

Viewed from another perspective, it should be recognized that in configurations according to the inventive subject matter the composition of the vapors from the storage tank is modified by mixing these vapors with a subcooled heavy hydrocarbon stream (the addition of heavy hydrocarbons increases the boiling point temperature, and therefore allows condensation of the mixture with LNG). This mixture is pumped to and separated in a downstream fractionator for recovery and recycling of the heavy hydrocarbons.

With further reference to Figure 2, vapor from the storage tank, stream 2, is split into stream 3 and stream 4. Stream 3 typically at a flow rate of 20 MMscfd is returned to the LNG ship via a vapor return line and return arm 54 for replenishing the displaced volume from ship unloading. Stream 4 typically at a flow rate of 20 MMscfd is mixed with the heavy hydrocarbon stream 16 (containing C₄ and heavier hydrocarbons). To raise the boiling point of the mixture, typically about 200 GPM to 500 GPM heavy hydrocarbons is required from the downstream fractionation system (the system is typically charged with the heavy hydrocarbons using an external source). The mixture is

cooled and condensed in exchanger 61 using the refrigeration content from the LNG stream 6 forming stream 18 typically at -240°F to -255°F .

It should be appreciated that the heavy hydrocarbon composition and flow rate can be adjusted in the fractionator as necessary to absorb the vapors from the storage tank during the ship unloading and the normal holding operation. Moreover, it should be recognized that the nature of the hydrocarbon is not critical so long as the hydrocarbon will increase the boiling point temperature. Therefore, suitable components for admixture with the vapor stream especially include propane, butane, and higher hydrocarbons. Alternatively, mixtures of hydrocarbons, chlorofluorocarbons and even non-hydrocarbon compositions may be employed so long as such components can be separated from the condensed vapor and increase the boiling point of the mixture.

In exchanger 61, stream 6 is heated from -255°F to about -240°F supply the necessary cooling for condensing the mixture 17. The condensate stream 18 is then pumped by pump 62 to about 120 psia to 170 psia forming stream 19. Prior to feeding the fractionator 64, the mixture is heated and partially vaporized in exchanger 63 to about -10°F to 150°F by cross exchange with the bottom liquid from the fractionator. The fractionation, operating at about 100 psia to 150 psia, separates the mixture producing an overhead liquid stream 22 (containing mostly C_2 and lighter components) and a bottom liquid stream 21 (containing mostly C_3 and heavier components). The fractionator is refluxed with the refrigeration from the LNG stream 17 in an overhead condenser 65 that is integral to the fractionator. Alternatively, the overhead condenser 65 can be located externally to the fractionator, and the liquid stream 22 can be separated in an externally located drum. The fractionator is reboiled using an external heat source stream 24 with a fired reboiler, steam or other heat source.

The overhead stream 22 which is depleted of the heavy hydrocarbons is mixed with the LNG stream 23 forming stream 10. The combined stream is pumped by the secondary pump 59 to typically 1000 psig to 1500 psig forming stream 11 which is then heated in LNG vaporizers to about 40°F to 60°F as needed to meet the pipeline

specifications. The LNG vaporizers are typically open rack type exchangers using seawater, fuel-fired vaporizers, or vaporizers using a heat transfer fluid.

Alternatively, as shown in Figure 3, the vapor from the storage tank is not returned to the ship, thereby eliminating the vapor return line and vapor return arm. Instead the vapor required by the ship for maintaining volumetric balance is generated with a small vaporizer located close to the ship. Here, a small stream 30 at a flow rate of 20 MMscfd is vaporized in heat exchanger 67 to produce vapor stream 3 that is used for replenishing the displaced volume from the ship. The heat source to the vaporizer can be seawater or ambient air. Such configurations are thought to result in significant cost savings in the terminal design particularly in facility where there is a great distance between the ship and the storage tank. Consequently, the entire vapor stream 2 from the tank is mixed with the heavy hydrocarbon stream 16, absorbed and condensed with LNG stream 6. Therefore, the flow rate of stream 16 is increased correspondingly to about 400 GPM to 1,200 GPM, as needed for the absorption of the higher vapor flow.

In yet another preferred aspect of the inventive subject matter, and especially where it is desired to extract LPG from the crude LNG or to otherwise modify the chemical composition of the LNG (e.g., to meet environmental regulations), additional cooling may be provided to the absorber as depicted in Figure 4. In such configurations, the overhead condenser 65 includes a second refrigeration coil 66 integral to the column that uses the high pressure LNG to provide additional cooling as needed for higher reflux duty required for LPG production. Alternatively, heat exchanger coil 66 and coil 65 can be located external to the column in separate heat exchangers, and liquid stream 22 can be separated in an external drum. Here, the LNG stream 26 exiting the condenser coil 65 at about -220°F to -240°F is split into two portions; stream 23 and stream 24. The split proportion varies between 0 to 100%, depending on the desirable LPG production; increasing stream 24 increases LPG production. With increasing LPG production, the distillate becomes leaner in composition resulting in the production of a LNG with less heating value that may be desirable for meeting environmental regulation.

Stream 24 is fed to the mid section of the fractionator which produces a bottom LPG stream 28, and an overhead distillate liquid stream 22 that is depleted of the heavy hydrocarbons. The distillate stream 22 is then mixed with the LNG stream 23 forming stream 10 typically at -220°F to -230°F that is further pumped by the secondary pump 59 to about 1,000 psig to 1,400 psig forming stream 11. The high pressure LNG stream is heated exchanged with the overhead vapor and is heated the reflux condenser coil 66 forming stream 27, typically at about -180°F to -200°F . Stream 27 is further heated in vaporizer 60 to meet the pipeline gas requirement. The bottom stream 28 is split into two portions; stream 25 and stream 21. Stream 21 is recycled back to exchanger 20 prior to be used for vapor absorption, and stream 25 can be sold as the LPG product.

Thus, specific embodiments and applications of LNG vapor handling and regasification have been disclosed. It should be apparent, however, to those skilled in the art that many more modifications besides those already described are possible without departing from the inventive concepts herein. The inventive subject matter, therefore, is not to be restricted except in the spirit of the disclosure. Moreover, in interpreting the specification, all terms should be interpreted in the broadest possible manner consistent with the context. In particular, the terms "comprises" and "comprising" should be interpreted as referring to elements, components, or steps in a non-exclusive manner, indicating that the referenced elements, components, or steps may be present, or utilized, or combined with other elements, components, or steps that are not expressly referenced.

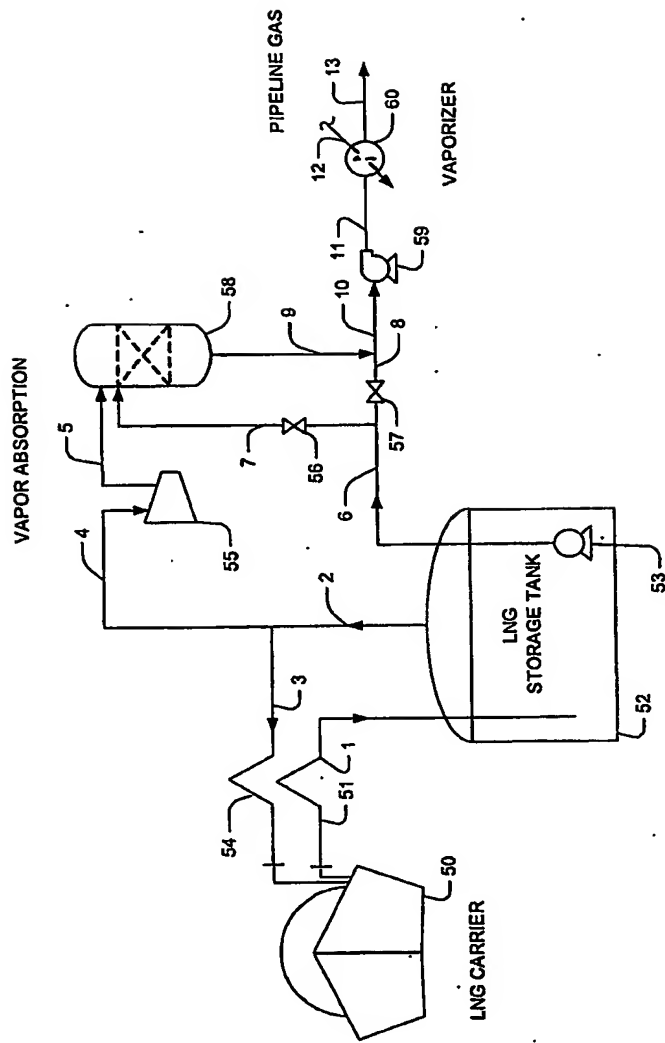


Figure 1 (Prior Art)

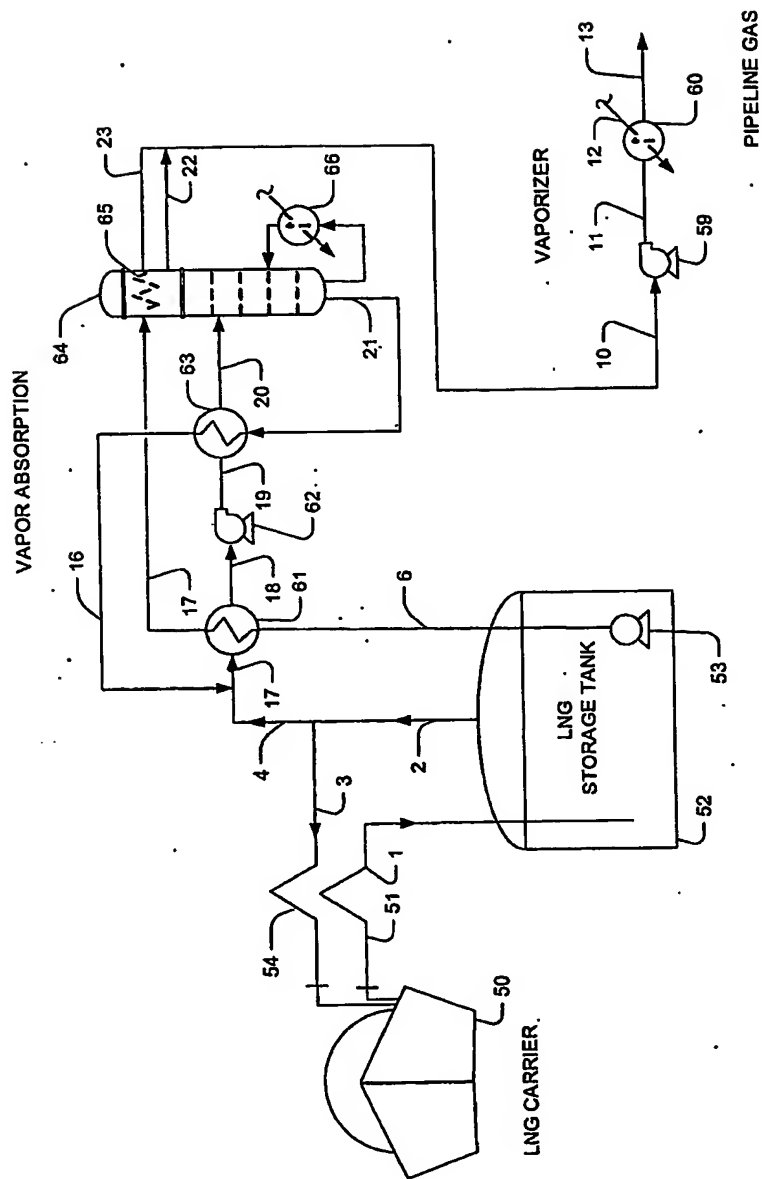


Figure 2

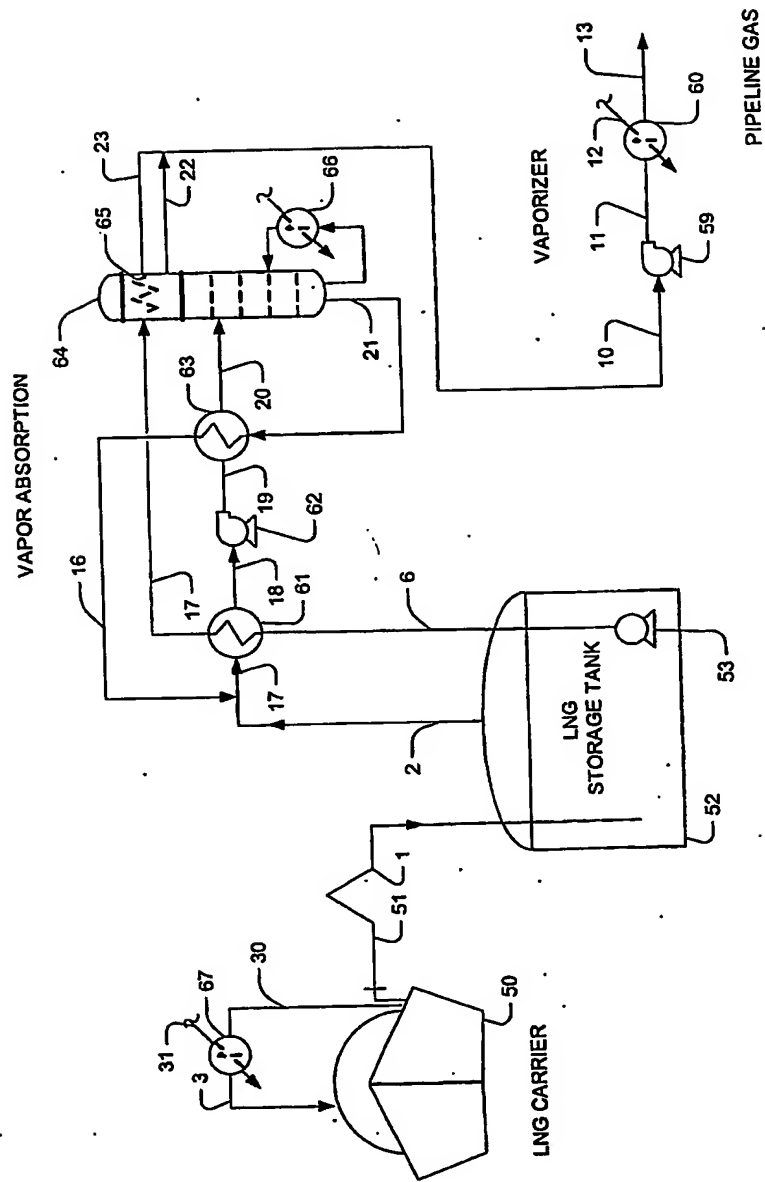


Figure 3

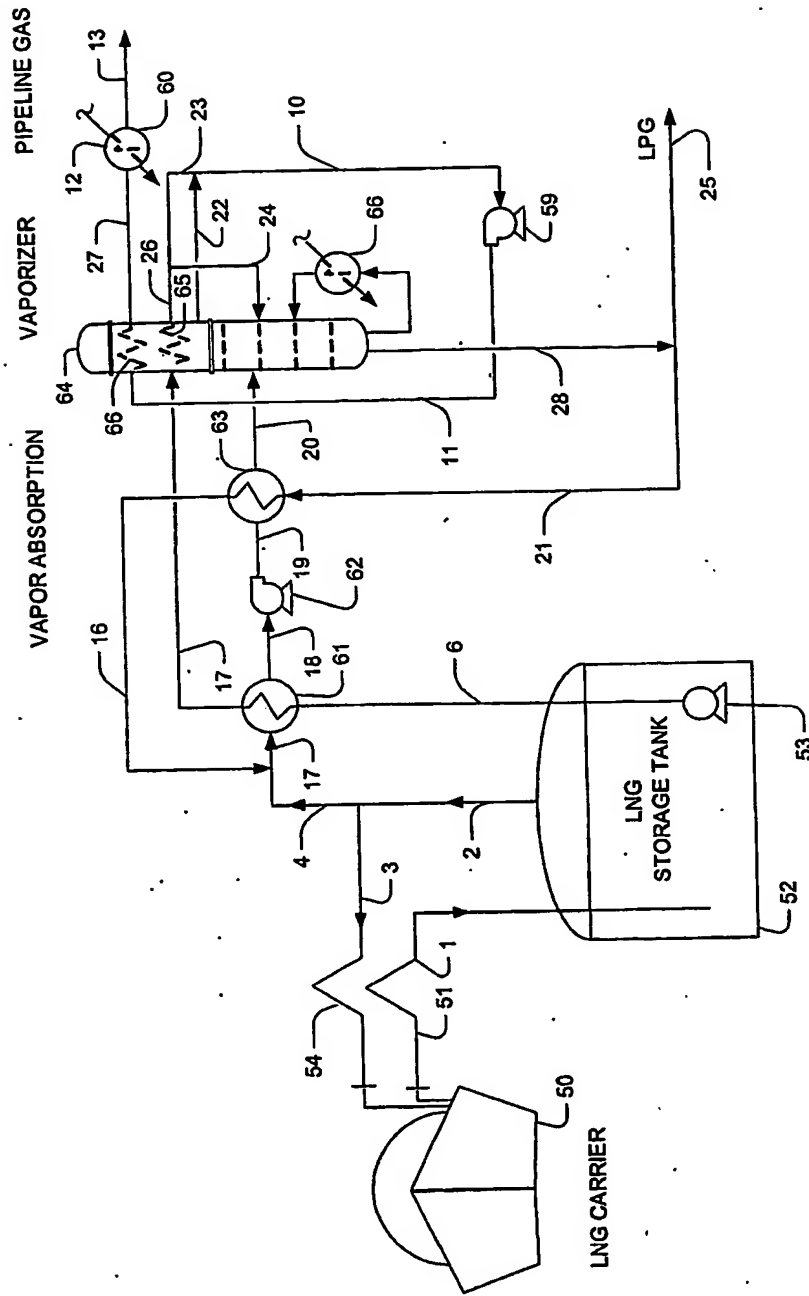


Figure 4